EFFECT ON NAPHTHA YIELD, OVERALL CONVERSION AND COKE YIELD THROUGH DIFFERENT OPERATING VARIABLES IN FCC UNIT USING ASPEN-HYSYS SIMULATOR

A thesis submitted in partial fulfillment of the requirements for the degree of

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by

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Rourkela

CERTIFICATE

This is to certify that the project report entitle "**Effect on naphtha yield, overall conversion and coke yield through different operating variables in FCC unit using Aspen-Hysys simulator"** submitted by **ANKIT KUMAR AGRAWAL (ROLL NO: 108CH030)** in the partial fulfillment of the requirement for the degree of the B.Tech in Chemical Engineering, National Institute of Technology, Rourkela is an authentic work carried out by him under my super vision. To the best of my knowledge the matter embodied in the report has not been submitted to any other university/institute for any degree.

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ABSTRACT

Fluid Catalytic Cracking Unit is the pump house of any refinery. Distillation is the initial step in the processing of crude oil and the residue which is coming out from the distillation column enters as the feed in the FCC unit. Gasoline is the main product of the FCC unit and it also produces byproduct which is more olefinic and hence more valuable. Simulation of the fractional distillation has been done to find out the feed composition which is the feed to the riser reactor. The unit was further simulated under the desired specifications to get the naphtha yield and compared with the plant data. Different graphs were plotted by varying feed temperature, flow rate, catalyst to oil ratio and were successfully compared with the modeled data. Further simulation was done with two regenerators and production of SOx was studied. The simulation result concludes that the SOx emission is lesser in case of one regenerator. Two sets of catalyst were chosen and the final yields were compared. Based on the plant requirement different types of catalyst are used. Finally the effect of riser height was studied in one riser and dual riser by keeping the operating parameters to be same and concluded with the fact that naphtha yield increases in case of dual riser.

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1. INTRODUCTION

A fluid catalytic cracking (FCC) unit converts low value heavy hydrocarbons having a carbon chain of more than 100 into valuable products gasoline and olefin compounds such as ethylene, propylene respectively. FCC riser reactor is designed to use acidic catalyst to decompose heavy oil, such as atmospheric gas oil (AGO and VGO), into more valuable lighter hydrocarbons at certain range. There can be further improvement of the products distribution in risers which can be made by changing the operating conditions [1, 2]. About 45% of worldwide gasoline production comes either directly from FCC units or indirectly from combination with downstream units, such as alkylation [3].

Earlier practices relied on thermal cracking which has now been completely replaced by fluidized cracking since it produces gasoline of higher octane number and also the by products which are more olefenic and hence more valuable. The light gases produced in the process contain more olefinic hydrocarbons than those by the thermal cracking process [4, 5]. The FCC unit mainly depends on circulating a zeolite catalyst, which is the main component, and accounts for around 26% with the vapour of the feed into a riser-reactor for a few seconds. The catalyst is circulated back into the regenerator where coke is burned and the catalyst is regenerated [6].

Due to the cracking reactions in the riser part some carbonaceous material such as coke gets deposited on the catalyst surface which reduces the activity of the catalyst so it is send back in the regenerator along with air. The cracking reaction is endothermic; the energy for which comes from the regenerator where catalyst is burned off in the presence of air which is an exothermic reaction. Some units of FCC are designed to use the supply of heat from the regenerator for the cracking purpose. These are known as “heat balance” units [7]. Petroleum Crudes consists of long chain of hydrocarbons which are processed through several separation processes like atmospheric distillation column, vacuum distillation column and finally oils of different boiling point ranges are obtained like gasoline (naphtha’s), diesel oil, LPG etc. Apart from these products, heavy oils (atmospheric gas oil or vacuum gas oil) are produced which have a boiling point of 343°C (650 °F) to 565°C (1050 °F). These heavy oils (AGO and VGO) are cracked in the FCC rector to form valuable petroleum products like gasoline LPG, lighter olefins. FCC unit is much preferred than the conventional thermal cracking process because it produces petroleum products of higher octane value.

As of 2006, FCC units were in operation at 400 petroleum refineries worldwide and about one-third of the crude oil refined in those refineries were processed in an FCC in order to produce high octane gasoline and fuel oils[8]. During 2007, the FCC units in the United States processed a total of 5,300,000 barrels (834,300,000 liters) per day of feedstock [9] and FCC units worldwide processed about twice that amount.

FCC units used in industries are usually of two types:

i. Side by side type and
ii. Stacked type reactor

In side by side reactor, which is used in this project for simulation purposes, reactor and regenerator are separated from each other and placed side by side. In case of stacked type reactor rector and regenerator are mounted together.

The basic process of FCC has got two major components i.e. reactor and regenerator. All the major processes happening here can be divided into following categories:

1.1. Preheat system

The feed in the FCC riser are the residue and the Atmospheric gas oil which comes out from the distillation column. The feed needs to be preheated before entering in the riser part. This is done by the feed preheat system which heats both the fresh and recycled feed through several heat exchangers and the temperature is maintained at about 500-700 °F. The gas oil consists of paraffinic, aromatics and naphthenic molecules and also contains various amounts of contaminants such as sulphur, nitrogen which have detrimental effect on the catalyst activity. Hence, in order to protect the catalyst feed pretreatment is essential which removes the contaminants and have better cracking ability thus giving higher yields of naphtha.

1.2. Riser

The riser is the main reactor in which most of the cracking reactions occur and all the reactions are endothermic in nature. The residence time in the riser is about 2–10 s. At the top of the riser, the gaseous products flow into the fractionator, while the catalyst and some heavy liquid hydrocarbon flow back in the disengaging zone. Steam is injected into the stripper section, and the oil is removed from the catalyst with the help of some baffles installed in the stripper [10]. The ideal riser diameter and length should be about 2 meters and 30 to 35 meters respectively.

1.3. Reactor

The earlier practice of carrying out the cracking reactions in the reactor has now been completely replaced by carrying out it in the riser part. This is done to utilize the maximum catalyst activity and temperature inside the riser. Earlier, no significant attempts were made for controlling the riser operations. But after the usage of the reactive zeolite catalyst the amount of cracking occurring in the riser has been enhanced. Now the reactor is used for the separation purpose of both the catalyst and the outlet products. Reactions in the riser are optimized by increasing the regenerated catalyst velocity to a desired value in the riser reactor and injecting the feed into the riser through spray nozzles.

The main purpose of reactor is to

i. Separate the spent catalyst form the cracked vapors and
ii. The spent catalyst flows downward through a steam stripping section to the regenerator.

The cracking reaction starts when the feed is in contact with the hot catalyst in the riser and continues until oil vapors are separated from the catalyst in the reactor separator. The hydrocarbons are then sent to the fractionator for the separation of liquid and the gaseous products. In the reactor the catalyst to oil ratio has to be maintained properly because it changes the selectivity of the product .The catalyst’s sensible heat is not only used for the cracking reaction but also for the vaporization of the feed. During simulation the effect of the riser is presumed as plug flow reactor where there is minimal back mixing, but practically there are both downward and upward slip due to drag force of vapor [11, 12].

1.4. Regenerator

The spent catalyst coming out from steam stripping section goes in the regenerator. Regenerator maintains the activity of the catalyst and also supplies heat to the reactor and therefore FCC unit is referred as Heat balanced unit [7]. Depending upon the feed stock quality there is deposition of coke on the catalyst surface. To reactivate the catalyst, air is supplied to the regenerator by using large air blowers. High speed of air is maintained in the regenerator to keep the catalyst bed in the fluidized state. Then through the distributor at the bottom air is sent to the regenerator. Coke is burned off during the process in significant amount. The regenerator operates at a temperature of about 715 °C and a pressure of about 2.41 bars. The hot catalyst (at about 715 °C) leaving the regenerator flows into a catalyst where any flue gases are allowed to escape and flow back into the upper part to the regenerator. The flow of the regenerated catalyst is regulated by a slide valve in the regenerated catalyst line. The hot flue gas exits the regenerator after passing through multiple sets of two-stage cyclones that removes entrained catalyst from the flue gas. The heat is produced due to the combustion of the coke and this heat is utilized in the catalytic cracking process. Heat is carried by the catalyst as sensible heat to the reactor. Flue gas coming out of the regenerator is passed through the cyclone separator and the residual catalyst is recovered. The specification of the catalyst will be discussed in detail at literature review. The regenerator is designed and modeled for burning the coke into carbon monoxide or carbon dioxide. Earlier, conversion of carbon to carbon monoxide was done which required lesser air supply hence the capital cost was reduced. But now a days air is supplied in such a scale that carbon is converted into carbon dioxide in this case the capital cost is higher but the regenerated catalyst has minimum coke content on it. The flue gases like carbon monoxide are burned off in a carbon monoxide furnace (waste heat boiler) to carbon dioxide and the available energy is recovered. The hot gases can be used to generate steam or to power expansion turbines to compress the regeneration air and generate power. There are two stage cyclones which remove any entrained catalyst from the flue gases.

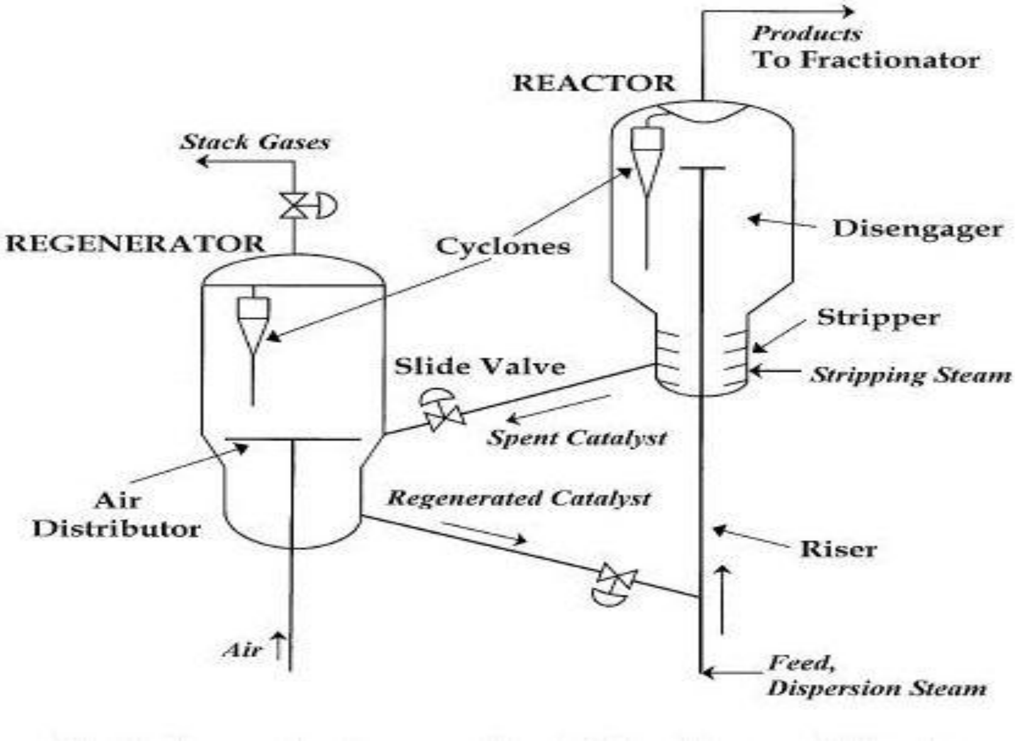


Figure 1: Schematic of the Fluid Catalytic Cracking Unit [13]

Simulation of the FCC reactor is done which is the objective of the project. The process parameters are varied at different conditions and the efficiency of the reactor is calculated. Simulation is done using Aspen Hysys. In the present simulation, the feed condition is obtained by simulating the atmospheric distillation column which is the input of FCC unit.

2. LITERATURE REVIEW

2.1. Pseudo-components

The pseudo components are used for the estimation of °API of the crude stream by characterizing the true boiling point of the crude. As the stream cannot be processed using 50-100 components in a refinery operation so the pseudo component concept is utilized. The crude oil is characterized into 30-40 components and its average properties can be used to represent the TBP and °API of the streams. The estimation is useful in evaluating the mass balances from volume balances. Generally in any refinery operation the flow rate is measured in barrels. So the flow rate can be converted to mass flow rate through the use of °API of the streams.

2.2. Riser Kinetics

There are various types of reactions taking place in Fluidized catalytic cracking, but the main reaction is the cracking of paraffin, naphthenic and side chain of aromatics. There are generally two types of reactions in FCC.

Primary Reactions

In this types of reactions primary cracking occurs through Carbenium ions in the following steps [14]

i) Formation of olefin by cracking of paraffin

$$nC_8H_{18} \rightarrow CH_4 + CH_3 - (CH_2)_4 - CH = CH_2$$

ii) Proton shift

$$CH_3 - (CH_2)_4 - CH = CH_2 + H_2O + \underset{\underset{\text{Zeolite}}{\overset{|}{O}}}{\overset{\overset{O}{|}}{Al}} - O - Si \rightarrow$$

$$\underset{\text{Carbenium ion}}{CH_3 - (CH_2)_4 - CH^+ - CH_3} + [HO - Al - Si]^-$$

iii) Beta Scission

Carbon–carbon scission takes place at the carbon in the position beta to the Carbenium ions and olefins.

$$CH_3 - (CH_2)_4 - CH^+ - CH_3 \rightarrow CH_3 - CH = CH_2 + CH_2^+ - CH_2 - CH_2 - CH_3$$

The newly formed carbenium ion reacts with another paraffin molecule which propagates the reaction. The reaction is terminated when the carbenium ion loses a proton to a catalyst and forms an olefin. Hydrogen transfer plays an important role in the FCC reactions since it decreases the olefinic product and converts it into more stable paraffin and aromatic rings.

$$\underset{\text{Olefin}}{3C_nH_{2n}} + \underset{\text{Naphthene}}{C_mH_{2m}} \rightarrow \underset{\text{Paraffin}}{3C_nH_{2n+2}} + \underset{\text{Aromatic}}{C_mH_{2m-6}}$$

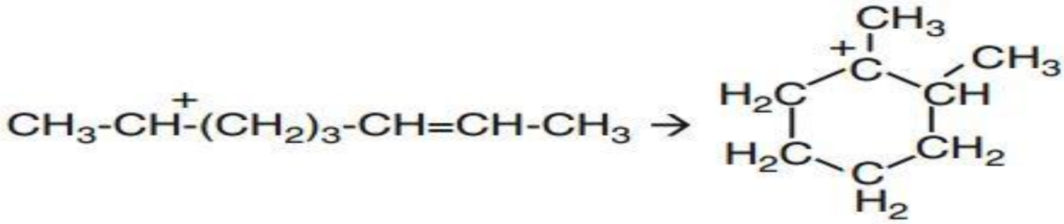
Secondary Reactions

The gasoline yield can be reduced due to the secondary reaction. The gasoline which is formed in the primary reaction can undergo secondary reaction through hydrogen transfer mechanism such as cyclisation, isomerization and coke formation.

i) Isomerization Reaction:

$$\underset{\text{Secondary Carbenium ion}}{R - CH_2 - CH_2 - CH^+ - CH_3} \leftrightarrows \underset{\text{Tertiary Carbenium ion}}{R - CH_2 - \overset{\overset{\displaystyle CH_3}{|}}{\underset{+}{C}} - CH_3}$$

ii) Cyclisation Reaction:



And this cyclisation reaction further cyclize to coke formation.

2.3. Catalytic activity

Commercial FCC catalysts are based on Y-zeolites as main component with ZSM-5 as additive [14]. There are three types of commercial catalyst:

i) Acid treated natural alumino-silicates
ii) Amorphous synthetic silica alumina combinations and
iii) Crystalline synthetic silica alumina catalysts called zeolites or molecular sieve [15].

The typical FCC catalyst consists of a mixture of an inert matrix (kaolin), an active matrix (alumina), a binder (silica or silica–alumina) and a Y zeolite. During the FCC process, a significant portion of the feedstock is converted into coke [16]. For the selectivity of the product zeolite is the essential part which ranges about 15 to 25 % of the catalyst and its structure is like tetrahedron with four oxygen atom at the corner and having an Aluminum or Silicon at the center. In general, the zeolite does not accept molecules larger than 8 to 10 nm to enter the lattice [17]. Matrix allows larger molecules of enter the lattice.

The use of ZSM-5 in FCC plants as an additive has also become very important in increasing both octane number and C3–C4 olefins [14].Y-zeolite is the active and the most important component in FCC catalysts. It provides the major part of the surface area and the active sites [18]. Thus, it is the key component, which controls catalyst activity and selectivity [19]. The catalytic activity of Y-zeolite is mainly controlled by its unit cell size (UCS) and to less extent by its crystal size. Recently, Al-Khattaf and de Lasa have studied the effect of Y-zeolite crystal size on the activity and selectivity of FCC catalysts [20, 21]. The conversion of coke and other catalytic activity depends on the acidic strength of the zeolite. So it is known that increase in the yield of coke occurs when there is high acidic strength (high UCS) value. High UCS also favors the hydrogen transfer reaction. As it is discussed the coke yield increases due to high UCS and it covers the active acidic part of the catalyst which decays the activity. Moreover the concept of octane number plays a vital part in selectivity of the reactor. That is why the hydrogen transfer reaction is an important one in the catalytic cracking reactor as it converts some of the light olefins into paraffins and aromatic compounds which have higher octane number value [22].

3. DESCRIPTION OF THE SIMULATION

3.1. ASPEN HYSYS

ASPEN HYSYS is a strong and versatile tool for the simulation studies, modeling and performance monitoring for oil and gas production, gas processing, petroleum refining, and air separation industries. It helps to check the feasibility of a process, to study and investigate the effect of various operating parameters on various reactions. It offers a high degree of flexibility because there are multiple ways to accomplish specific tasks. This flexibility combined with a consistent and logical approach to how these capabilities are delivered makes HYSYS an extremely versatile process simulation tool. The usability of HYSYS is attributed to the following four key aspects of its design:

i) Event Driven operation
ii) Modular Operations
iii) Multi-flow sheet Architecture
iv) Object Oriented Design

3.2. FCC and ASPEN HYSYS

The FCC unit works through various cracking reaction in the riser reactor section of this unit. Different types of model of the FCC reactors are available in ASPEN HYSYS such as:

i) One riser
ii) Two riser
iii) Risers with mid-point injection
iv) One stage regenerator
v) Two stage regenerator(flue gas in series)
vi) Two stage regenerator(separate flue gas)

4. PROBLEM DESCRIPTION & SIMULATION

4.1. PROBLEM

The present simulation is done to study the effects of various operating and design conditions on

i) Naphtha yield
ii) Coke yield
iii) Total conversion

Here, the variation in the yield pattern is studied using the following model keeping the designing parameters same in all the models:

i) One riser
ii) Dual riser
iii) Two stage regenerator (Flue gas in series)

Finally, the results of simulation are compared with the plant data of Qianguo Petroleum Refinery.

4.2. SIMULATION

As mentioned above the main purpose of the present work is to study the effects of variation of process conditions on the production of naphtha yield in the FCC. For the present study, a refinery process was simulated in order to assist in the simulation. The details are discussed below:

4.2.1. Process Flow Diagram

To represent the refinery process + FCC unit in Aspen HYSYS, the first step is to make a process flow diagram (PFD). In Simulation Basic Manager, a fluid package was selected along with the components which are to be in the input stream. In the process, Peng-Robinson was selected as the fluid package as it can handle hypothetical components (pseudo-components).

The non-oil components used for the process were H_20, C3, i-C4, n-C4, i-C5 and n-C5. The pseudo-components were created by supplying the data to define the assay. The fluid package contains 44 components (NC: 44): 6 pure components (H_2O plus five Light Ends components) and 38 petroleum hypocomponents). In order to go to the PFD screen of the process the option "Enter to simulation Environment" was clicked on. An object palette appeared at right hand side of the screen displaying various operations and units.

The PFD of the process is given below:

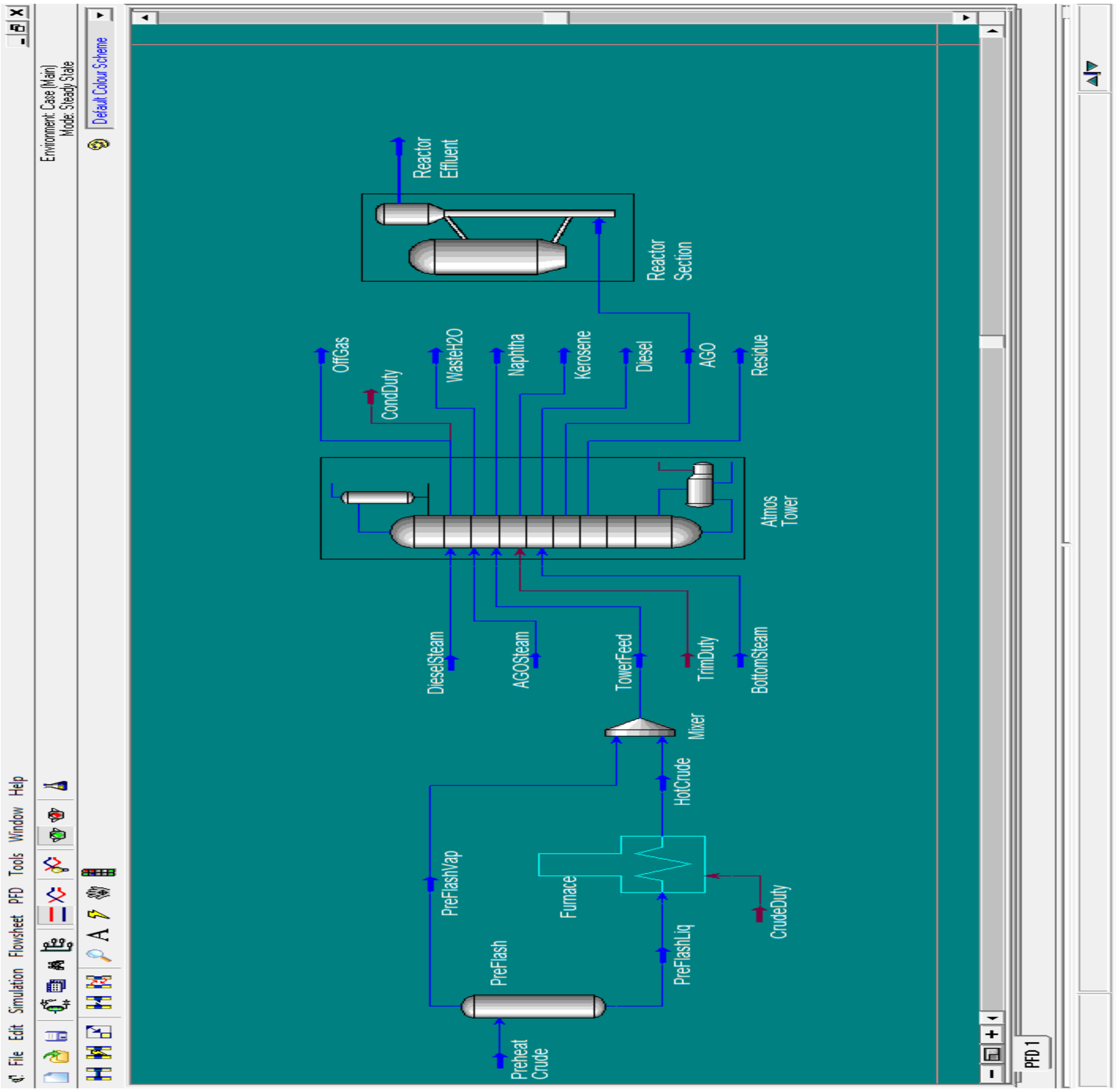


Figure 2: PFD of the simulation carried out in ASPEN HYSYS

Here,

PreFlash is a separator.

Furnace is a heater.

Mixer is a mixer.

Atmos Tower is a distillation column operated at 1 atm.

Reactor Section is the FCC Unit in which AGO (Atmospheric Gas Oil) is used as the feed.

4.2.2. Process Description

The Crude Oil enters the PreFlash unit, a separator used to split the feed stream into its liquid and vapour phases at 450 F and 75 psia having a molecular weight of 300 and °API of 48.75. The crude stream separates into the PreFlashVap and PreFlashLiq consisting of purely vapour and liquid respectively. The PreFlashLiq enters the crude furnace flashing part of the liquid to vapour which comes out as stream, HotCrude having a temperature of 650 F. The PreFlashVap and HotCrude streams are then inlet into the Mixer resulting into the formation of the TowerFeed. The Atmos Tower is a column having Side Stripper systems to draw out Kerosene, Diesel and Atmospheric Gas Oil. Naphtha is drawn from the condenser and Residue from the reboiler. The Atmospheric Gas Oil (AGO) is then used as the feed to the Reactor Section, the FCC unit. The FCC Unit was configured to have one or two risers with the geometry as per the data collected by Derouin [23]. It was assumed that no heat loss occurs in the FCC unit. Catalyst was decided upon and operating conditions were set.

Results were noted for the variation of Naphtha Yield, Coke (wt. %) and Total conversion with change in the following operating conditions:

i) C/O ratio
ii) Feed Flow Rate
iii) Feed Temperature
iv) Reactor Temperature

Total conversion is attributed to the conversion of the feedstock to the FCC into H_2S, Fuel Gas, Propane, Propylene, n-Butane, i-Butane, Naphtha, Butenes and Coke while the conversion of feedstock to Light Cycle Oil and Bottoms is not considered in the calculation of total conversion.

4.2.3. Components

Description of various components used in the PFD and the conditions at which they are operated are described here:

i) Separator (PreFlash)

No heat loss was assumed for the separator of volume 70.63 ft^3. Preheat Crude entered at 450 F and 75 psia with a 100,000 barrels/day flow rate containing mostly liquid. It had a molecular weight of 300 and API Gravity of 48.75. The Preheat Crude was separated into PreFlashLiq (450 F, 75 psia) and PreFlashVap (450°F, 75 psia).

ii) Heater (Furnace)

No heat loss was assumed for the Heater. PreFlashLiq entered the furnace at 450 F and 75 psia. Its main purpose was to partially vaporize the feed and increase its temperature to the feed conditions needed for the distillation column. The outlet stream hot crude had conditions 650°F, 65 psia.

iii) Mixer (Mixer)

The main purpose of the Mixer was to mix two streams, HotCrude (650 F, 65 psia) and PreFlashVap (450°F, 75 psia) to give on stream, TowerFeed (641.5°F, 65 psia) which is the feed stock to the distillation column.

iv) Distillation Column (Atmos Tower)

The feed to the column enters at 641.5°F, 65 psia. The column separates the feed into six fractions namely: Off Gas, Naphtha, Kerosene, Diesel, Atmospheric Gas Oil and Residue. The main column consists of 29 trays.

v) Fluidized Catalytic Cracking Unit (Reactor Section)

The Atmospheric Gas Oil was taken as the feed for this Unit. Initial conditions are given in the appendix attached. Results are shown in the Results and Discussion section.

The simulation for the FCC unit needs simulated feedstock. For the feedstock for the FCCU, Crude Petroleum, data was obtained from ASPEN HYSYS. The feed of molecular weight 300 and API Gravity 48.75 was used at a temperature of 450 °F and pressure of 75 psia.
Given below are the properties used for the crude petroleum feedstock:

Table 1: Crude Petroleum Simulation Feedstock Properties

Preheat Crude (Feedstock)	
Temperature [°F]	450
Pressure [psia]	75
Liquid Volume Flow [barrels/day]	100000

Table 2: Bulk Crude Properties

Bulk Crude Properties	
MW	300.00
API Gravity	48.75

Table 3: Light Ends Liquid Volume Percent of Crude Petroleum Feedstock

Light Ends Liquid Volume Percent	
i-Butane	0.19
n-Butane	0.11
i-Pentane	0.37
n-Pentane	0.46

Table 4: API Gravity Assay of Crude Petroleum Feedstock

API Gravity Assay	
Liq Vol% Distilled	**API Gravity**
13.0	63.28
33.0	54.86
57.0	45.91
74.0	38.21
91.0	26.01

Table 5: Viscosity Assay of Crude Petroleum Feedstock

Viscosity Assay		
Liquid Volume Percent Distilled	**Viscosity (cP) 100°F**	**Viscosity (cP) 210°F**
10.0	0.20	0.10
30.0	0.75	0.30
50.0	4.20	0.80
70.0	39.00	7.50
90.0	600.00	122.30

Table 6: TBP Distillation Assay of Crude Petroleum Feedstock

TBP Distillation Assay		
Liquid Volume Percent Distilled	**Temperature (°F)**	**Molecular Weight**
0.0	80.0	68.0
10.0	255.0	119.0
20.0	349.0	150.0
30.0	430.0	182.0
40.0	527.0	225.0
50.0	635.0	282.0
60.0	751.0	350.0
70.0	915.0	456.0
80.0	1095.0	585.0
90.0	1277.0	713.0
98.0	1410.0	838.0

The simulation was done and the product properties for the Atmospheric Distillation Tower were obtained. The Distillation Tower had six outlets out of which the top gaseous product stream had no mass flow. Hence only properties for the five outlet streams which consisted of Naphtha, Kerosene, Diesel, Atmospheric Gas Oil (AGO) and Residue were obtained. The AGO stream was then used in a 1-riser FCC unit to obtain the Naphtha Weight percentage and total conversion by varying different parameters such as Catalyst to oil ratio, feed temperature, feed flow rate and riser height. . The conditions under which the FCC unit was operated are given in Appendix 1.

Table 7: Atmospheric Distillation Tower Product Properties

Atmospheric Distillation Tower Product Properties					
Product Name	**Liquid Volume Flow [barrels/day]**	**Molecular Weight**	**Mass Density [API]**	**Temperature [°F]**	**Pressure [psia]**
Naphtha	20000	138.4	86.12	163.9	19.7
Kerosene	13000	210.1	118.8	449.2	29.84
Diesel	16998	289.1	109.6	478.4	30.99
AGO	5017	390.1	114.6	567.2	31.7
Residue	41322	614.6	83.21	657.1	32.7

5. RESULTS AND DISCUSSION:

The following table depicts the specification in which simulation was carried out and compared with the plant data (Qianguo Petroleum Refinery) result [24, 23].

Table 8: Design parameters

Specification	Simulation Data Value	Plant data value
Height	32m	36.2m
Diameter	1m	0.8m
Flow Rate	85kg/sec	25.52kg/sec
Feed Temperature	650K	463.2K
Catalyst to oil Ratio	5.53	6.30

On simulation of the FCC unit under the above stated conditions the following outputs have been obtained in terms of weight %.

Table 9: Outlet Composition Results from FCC simulation

COMPONENTS	WEIGHT (%)
H_2S	1.2508
FUEL GAS	3.5345
PROPANE	2.1537
PROPYLENE	4.2208
N-BUTANE	1.3596
I-BUTANE	2.9359
NAPHTHA	35.0832
BUTENES	5.6542
LCO	18.4137
BOTTOMS	21.5850
COKE YIELD	3.8086
CONVERSION	60.0013
TOTAL	100

The simulated results were compared with the plant data result of Naphtha and Coke yield:

Table 10: Comparison of the simulation results with the plant data result

COMPONENTS	Simulation Result Weight (%)	Plant data Result (Weight %)
NAPHTHA	35.0832	48.90
LCO	18.4137	21.74
COKE YIELD	3.8086	8.28
CONVERSION	60.0013	72.47

The Naphtha coming out from the plant data is more than the simulated data due to the difference in the operating parameters. The catalyst used in the simulation is Conquest 95 and the composition of the catalyst is different as used in the plant data. As height increases the residence time in the reactor increases this leads to more cracking of the feed and hence more gasoline yield as in case of simulation result.

5.1. EFECT OF FEED TEMPERATURE

The simulation was done by using different values of feed temperature which resulted in different yield of naphtha and overall conversion. As the temperature of the feed rises from a certain value naphtha yield decreases slightly and so is the total conversion. This is because there is not enough cracking reaction in the riser reactor in presence of the catalyst. Cracking would start before the riser which would decrease the percentage yield of the product.

Table 11: Variation of naphtha & coke yield, total conversion with feed temperature

FEED TEMPERATURE (^{0}F)	NAPHTHA (WT %)	TOTAL CONVERSION (%)	COKE YIELD (WT %)
386	43.6586	79.9801	6.2531
392	43.62	79.92	6.2259
398	43.598	79.8668	6.1985
402	43.5668	79.8092	6.1709
410	43.5351	79.7511	6.1433

5.2. EFFECTS OF C/O RATIO

Simulation is done by changing the catalyst to oil ratio and the effect is studied on gasoline and coke yield. The naphtha yield increases with the increasing C/O ratio however, the rate of increase in the naphtha yield decreases at higher values of C/O ratio. This can be attributed to the fact that at substantially high catalyst concentration cracking of pseudo components in the naphtha range (known as secondary cracking reactions) also increases which causes a decrease in the rate of increase of naphtha yield with C/O ratio. On the other hand, the increasing C/O ratio leads to increase in catalyst concentration, and hence increase in rate of both primary and secondary cracking. This increases overall number of moles cracked on the catalyst surface and hence increases amount of coke deposited on the catalyst. As in the modeled data the Catalyst to oil ratio is more than the simulation data so more cracking reactions takes place which increases the naphtha yield.

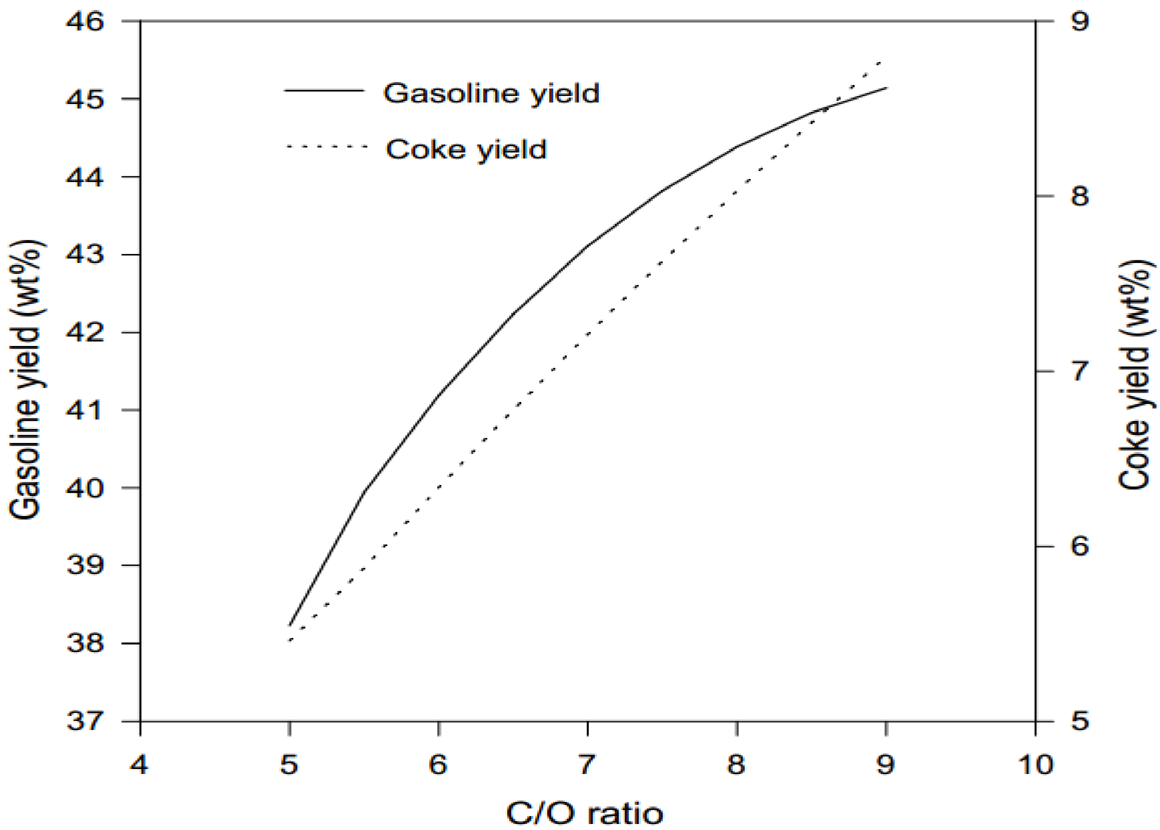


Figure 3: Graph of Naphtha Yield and coke yield vs. C/O Ratio [25]

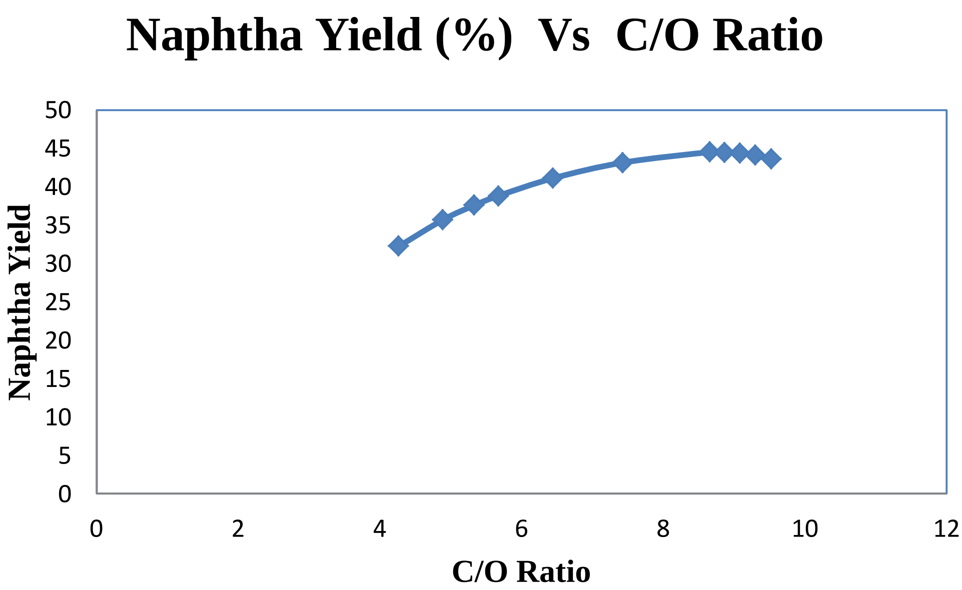


Figure 4: Graph of Naphtha Yield vs. C/O Ratio

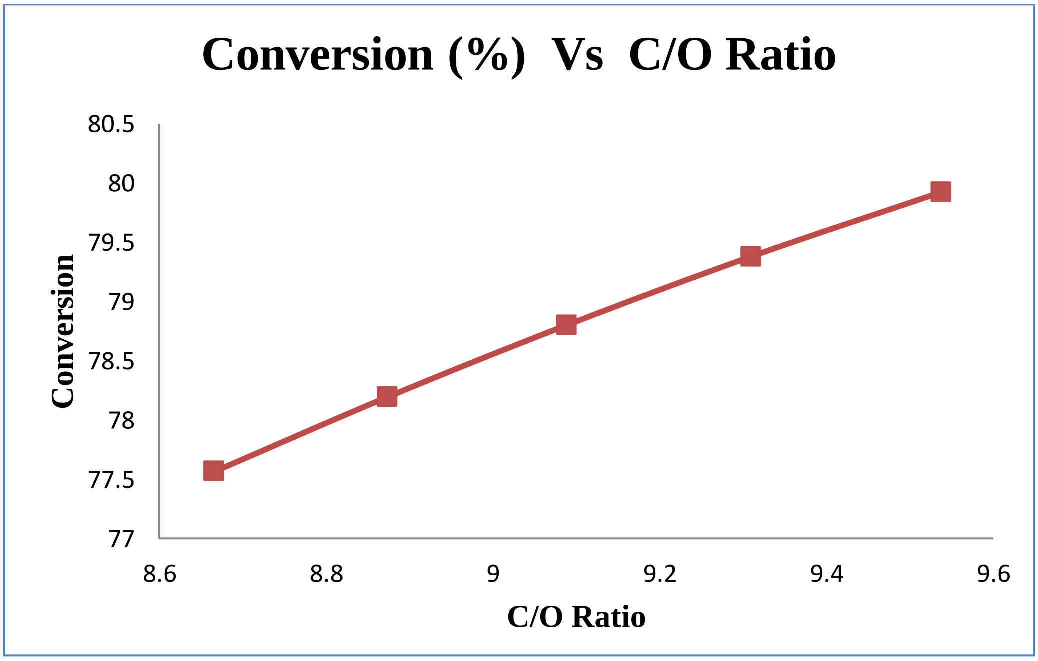


Figure 5: Graph of Conversion % vs. C/O Ratio

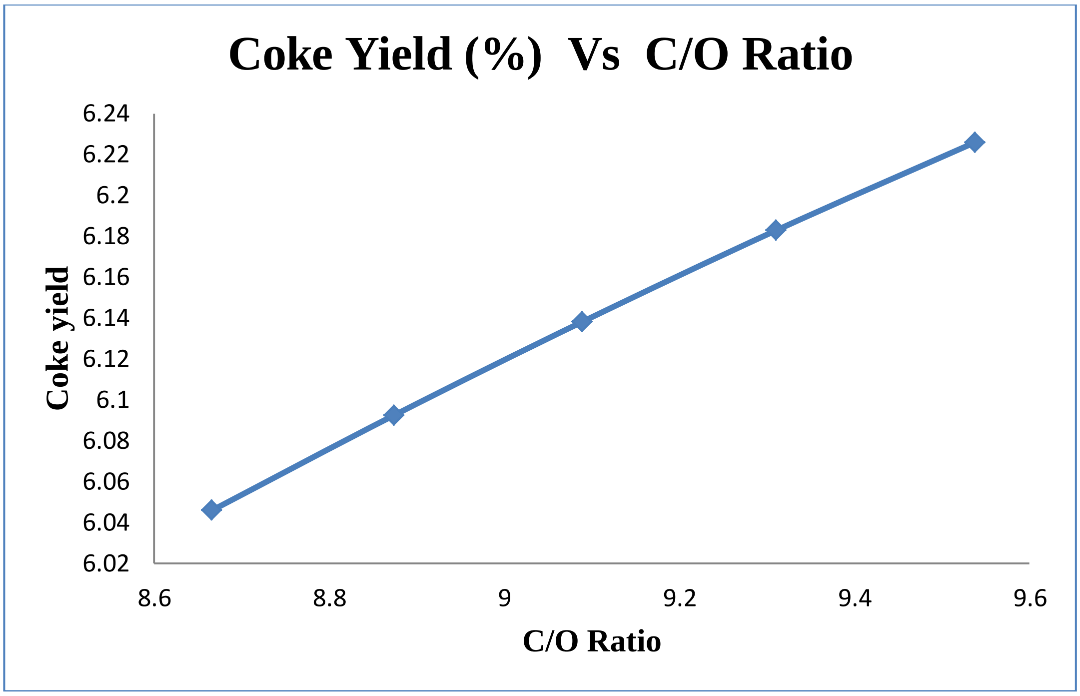


Figure 6: Graph of Coke Yield % vs. C/O Ratio

5.3. EFFECT OF FLOWRATE

Increasing the flow rate of the feed oil to the riser first increases the naphtha yield to a certain point and further increase in the feed oil decreases the naphtha yield as shown in the following graph. As flow rate of the feed oil to the riser increases, first the naphtha yield increases to a certain point and further increasing the flow rate yield decreases as shown by the graph below. This is because ,with high flow rate riser time decreases resulting less yield of naphtha; and then decreasing flow rate riser time increases which results to more yield. After a certain flow rate the riser time becomes very high resulting more cracking of naphtha to lighter components .but the total conversion increases with increase of the riser time.

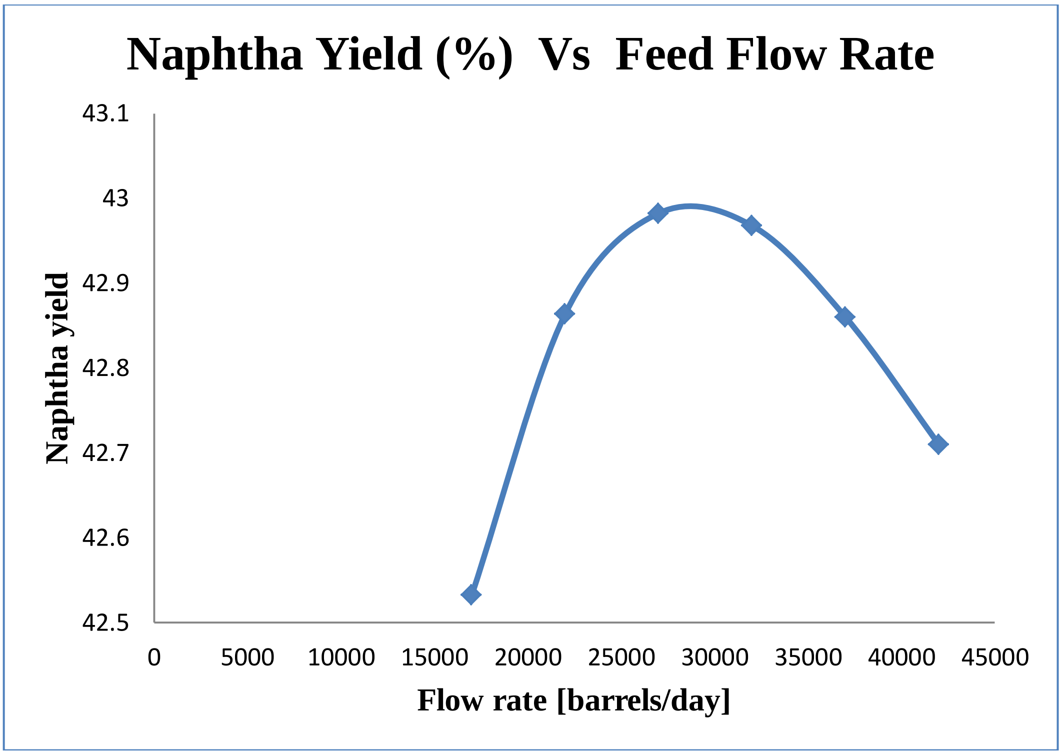


Figure 7: Effect on Naphtha Yield % vs. Feed Flow Rate

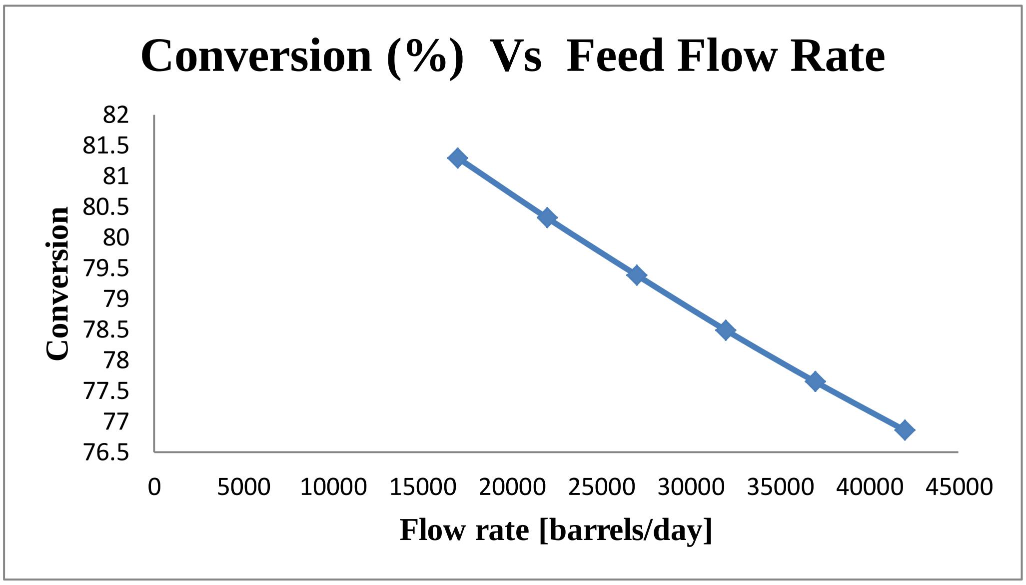


Figure 8: Effect on total Conversion % vs. Feed Flow Rate

5.4. COMPARISON OF ONE RISER AND DUAL RISER

Simulation was done using conquest type catalyst (zeolite 24.38 %) in two types of riser reactor i.e. one riser reactor and dual riser reactor at process condition as follows: [23]

Table 12: Specification data used for the comparison of one riser and dual riser

Specification	Simulation Data Value
Height	32m
Diameter	1m
Mass Flow Rate	85kg/sec
Feed Temperature	650K
Catalyst to oil Ratio	5.53
Catalyst used	Conquest 95
Reactor Plenum Temperature	833K

Table 13: Comparison of simulation data between one riser and two risers at given conditions

Component	One riser	Dual riser
H_2S	1.2411	0.3004
FUEL GAS	2.4126	1.7184
PROPANE	1.2549	0.7704
PROPYLENE	2.8034	3.2668
N-BUTANE	1.1734	0.8067
I-BUTANE	2.8034	1.6026
BUTENES	3.8724	4.7597
NAPHTHA	36.5292	38.7242
LCO	19.9435	18.7605
BOTTOMS	25.128	25.4484
COKE YIELD	3.5712	3.8420
TOTAL	100	100
CONVERSION	54.9285	55.7912

As shown in the Table 13, the gasoline yield is more in case of dual riser reactor (38.75% as compared to 36.52% of one riser). The overall conversion and coke yield also increases in the process.

Table 14: Simulation data of one riser reactor using AF3 Catalyst

COMPONENTS	PERCENTAGE (%)
H2S	1.2717
FUEL GAS	3.6339
PROPANE	2.2100
PROPYLENE	4.2968
N-BUTANE	1.3767
I-BUTANE	2.9855
BUTENES	5.7392
NAPHTHA	35.2324
LCO	18.1032
BOTTOMS	21.2281
COKE YIELD	3.9225
TOTAL	100
CONVERSION	60.6687

Using the same process condition and design parameter simulation of one riser reactor has been done by using two sets of catalyst (see tabulated results of table 10 &11). The catalyst used is A/F3 and conquest95 catalyst .The detailed composition is shown in the appendix. Mainly in a catalyst zeolite is the most important factor as it characterizes the selectivity of the process. Both A/F3 and conquest have zeolite concentration of 26.69% and 24.38 %. About 20-25% zeolite concentration is good for gasoline yield. More than that results over-cracking of the feed resulting lighter olefins which is observed in the case of A/F3 catalyst (ex. Propylene conc. 4.29% in case of AF3). As more coke yield and olefins yield occur when A/F3 is used, so the total conversion also increases. But when conquest 95 catalyst is used gasoline production is more as compare to A/F3 process (36.5292% whereas in case of A/F3 35.2324%). The simulated result shows that light paraffin's like N-butane and iso-butane production is more .this shows that the gasoline product of this process has high octane value as paraffin's and aromatics are good anti-knocking agents undergoing hydrogen transfer mechanism. So catalyst have different objective, one increases the oil quality and the second increases the gasoline yield.

5.5. EFFECTS OF FLOW RATE IN BOTH REACTORS:

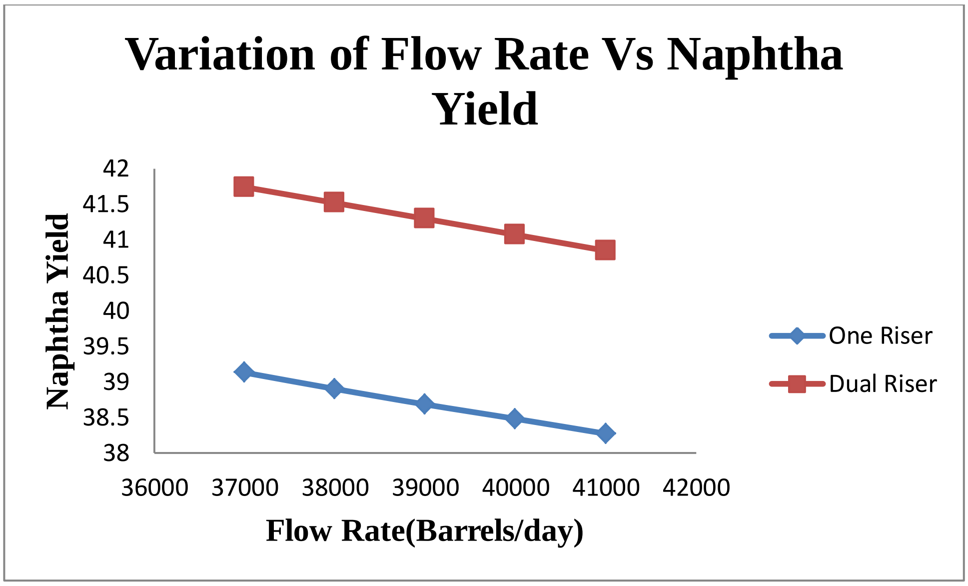


Figure 9: Effect of naphtha yield vs. flow rate

If we have to maintain maximum flow rate and we have to increase the residence time of the reactor instead of changing the riser height dual riser reactors are used in which the stream is divided into two and the flow rate is divided in each riser. Due to high flow rate the reaction time in the reactor will be very less, so very less time will be there for efficient contact between catalyst and feed and the naphtha yield decreases as the flow rate increases. At the same flow rate the dual riser shows higher yield than one riser reactor because in case of dual riser the flow rate is divided into two streams, so flow rate will be half and the feed velocity in the riser will be less. So there is efficient time for the cracking process which will result in more gasoline yield.

5.6. EFFECT OF RISER HEIGHT

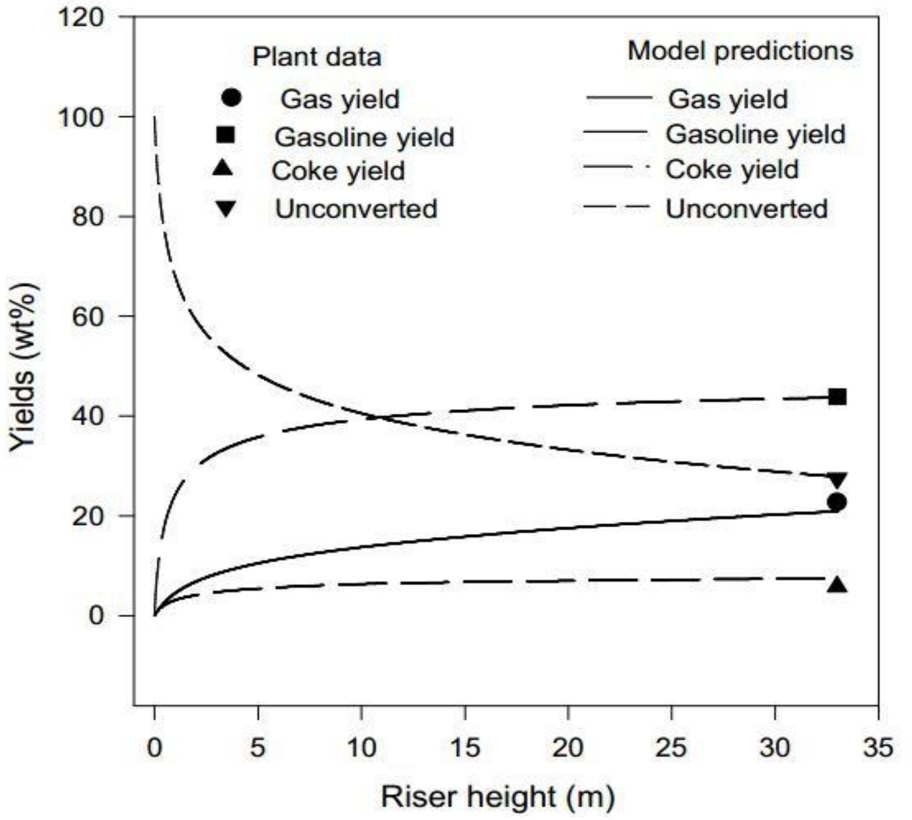


Figure 10: Effect of riser height on different yield [25]

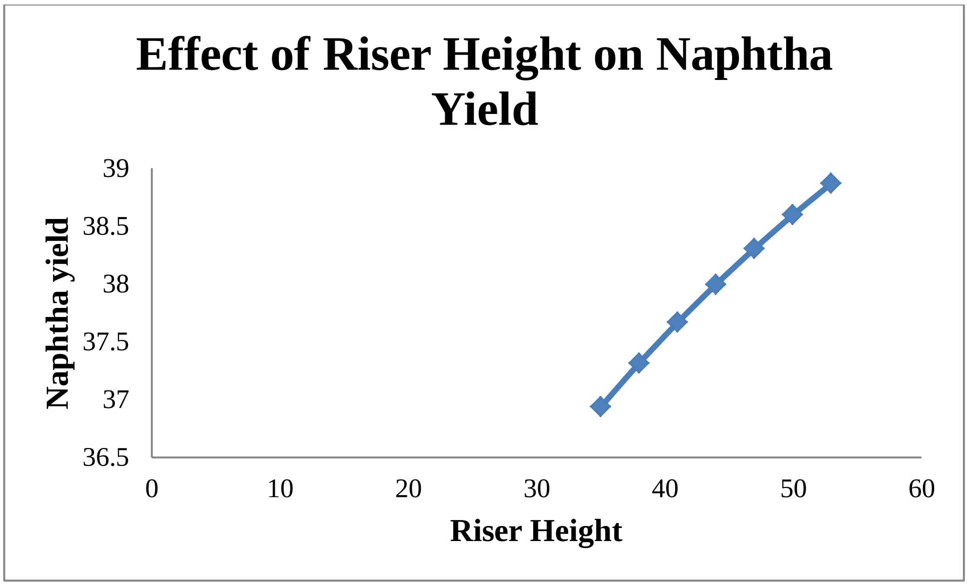


Figure 11: Effect of riser height on Naphtha yield

As shown, naphtha yield will increase as height increases. First it will increase rapidly but as the height goes on increasing the increase in naphtha yield decreases which is attributed with the plant result. As height increases at first the residence time in the reactor increases .this leads to more cracking of the feed .but when height is further increased secondary cracking dominates the process and naphtha yield decreases. In the figure 12 the naphtha yield is still increasing as height increases because the flow rate is maintained at 85kg/sec .At this flow rate there is minimum residence time in the reactor, so naphtha yield is increasing as height reaches about 60 meters. It can be shown in the table 11 that in case of dual riser at 32 meter height and with the same process condition the yield is about 38.72% which is 36.8% in case of single riser.

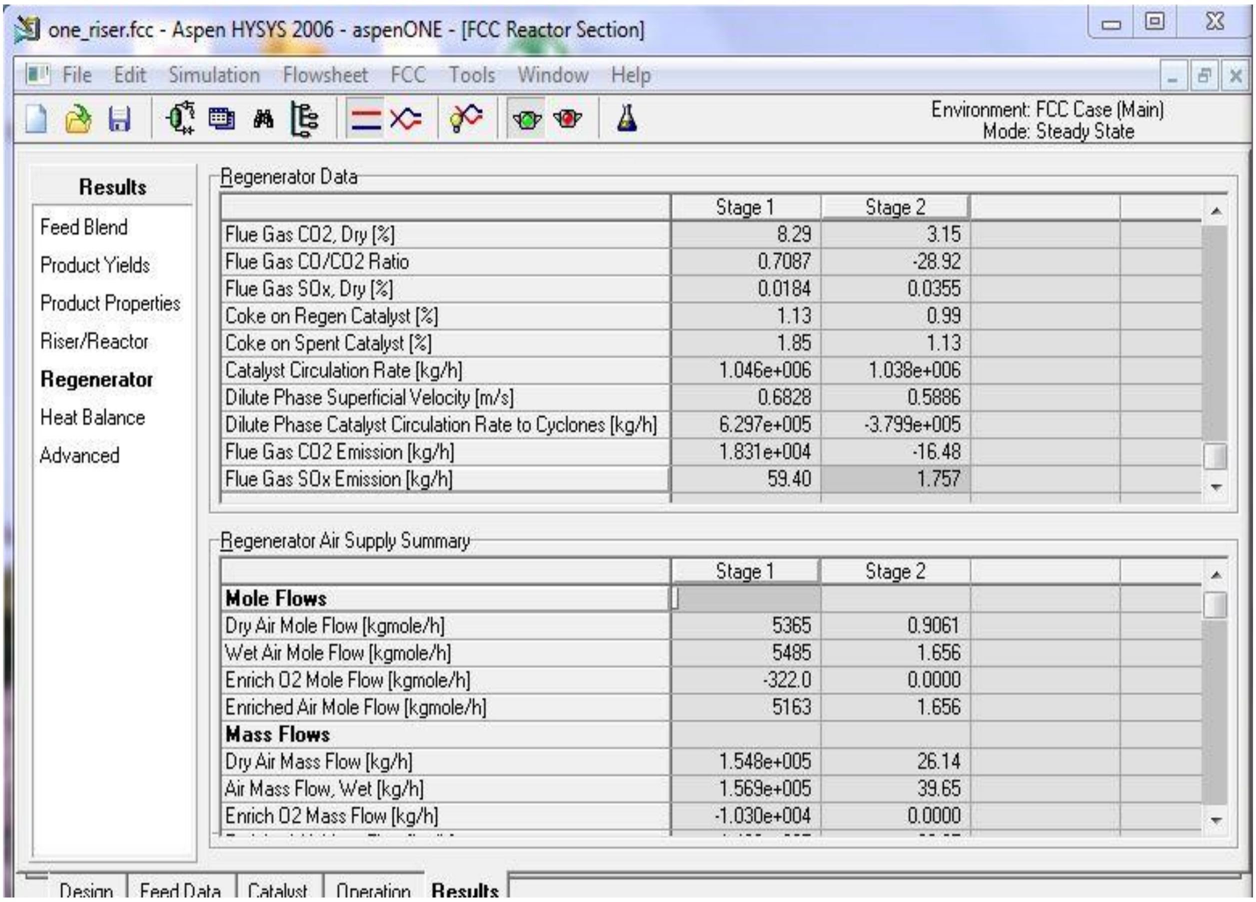
5.7. TWO STAGE REGENERATOR (FLUE GAS IN SERIES)

During the combustion process and from the carryover of catalyst particles atmospheric contaminants are formed in the regenerator. Among many contaminants SO_x is the major contaminant which has very detrimental effect on the environment. Sulfur trioxide can constitute up to about 10% of the total $S0_2$ (sulfur dioxide) plus $S0_3$, compared to a typical combustion effluent with $S0_3$ at a nominal 1-3% [26].

The presence of SO_3 in the flue gas can also lead to the formation of sulfuric acid. If the flue gas temperature falls below the sulfuric acid dew point (150-175°C, 303-347°F),[27] SO_3 and water (H_2O) will condense out to form the acid and corrosion of downstream equipment may result.

Catalyst activity will also be reduced and hence percentage yield will be reduced. Two regenerators are used which will further reduce the SO_x emission and increase the percentage yield. In the first stage partial combustion takes place and the spent catalyst goes in the second regenerator and complete combustion takes place in presence of air and therefore the catalyst activity is enhanced by minimizing coke formation.

SO_x emission causes a wide range of environmental and health problems in the way it reacts with oxygen. The impacts include respiratory problems and also lead to acid rain which has detrimental effect on the historic monuments. Two stage regenerator is used for the simulation having the same operating conditions. A decrease in the SO_x emission is noted as in case of one – stage regenerator.



Regenerator Data

	Stage 1	Stage 2
Flue Gas CO2, Dry [%]	8.29	3.15
Flue Gas CO/CO2 Ratio	0.7087	-28.92
Flue Gas SOx, Dry [%]	0.0184	0.0355
Coke on Regen Catalyst [%]	1.13	0.99
Coke on Spent Catalyst [%]	1.85	1.13
Catalyst Circulation Rate [kg/h]	1.046e+006	1.038e+006
Dilute Phase Superficial Velocity [m/s]	0.6828	0.5886
Dilute Phase Catalyst Circulation Rate to Cyclones [kg/h]	6.297e+005	-3.799e+005
Flue Gas CO2 Emission [kg/h]	1.831e+004	-16.48
Flue Gas SOx Emission [kg/h]	59.40	1.757

Regenerator Air Supply Summary

	Stage 1	Stage 2
Mole Flows		
Dry Air Mole Flow [kgmole/h]	5365	0.9061
Wet Air Mole Flow [kgmole/h]	5485	1.656
Enrich O2 Mole Flow [kgmole/h]	-322.0	0.0000
Enriched Air Mole Flow [kgmole/h]	5163	1.656
Mass Flows		
Dry Air Mass Flow [kg/h]	1.548e+005	26.14
Air Mass Flow, Wet [kg/h]	1.569e+005	39.65
Enrich O2 Mass Flow [kg/h]	-1.030e+004	0.0000

Figure 12: Simulation result of a two stage regenerator.

It has been observed that the coke yield is less in two-stage regenerator so there is an increase in the rate of the cracking reaction. This increases the naphtha yield and overall conversion.

6. CONCLUSION

The FCC unit was simulated to obtain the final yields which were compared with the plant data. The Naphtha yield from the present simulation comes out to be 35.0832% while the same is 48.9% in plant data. This difference can be attributed to different operating parameters such as catalyst to oil ratio, feed flow rate and riser temperature etc.

In the present simulation atmospheric gas oil has been taken as the feed to the FCC unit and the processing conditions such as flow rate, C/O ratio, feed temperature were varied to observe the operation of the FCC unit. Further these results were compared with the modeled output. The overall yield obtained by using different sets of catalyst (A/F3 and Conquest 95) was also calculated. The difference in the yield is due to the different compositions of the catalyst which has been precisely mentioned in the Appendix (d, e). The yield while using the A/F3 catalyst is lesser than that using Conquest 95. But the octane number of the oil obtained is higher than that in Conquest 95. So it can be concluded that the selectivity of the catalyst depends entirely upon the process plant and accordingly catalysts are used. From the various graphs it is seen that there is an optimum condition for each process and the plants should run by it to get the maximum output.

The yield percentage in case of one riser and dual riser reactor is also obtained and it was found that it is more in case of dual riser. Further two regenerators FCC model was used and it was found that unit the SO_x emission to the atmosphere was lesser than the one regenerator. Using two stage regenerator SO_x emission is reduced to (1.757kg/hr.) while using one stage regenerator it was (59.40kg/hr.). Due to the complete combustion in case of two stage regenerator the catalytic activity is enhanced and produces high yield of naphtha.

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Appendix

a) One Riser

LEGENDS Calgary, Alberta CANADA	Case Name: ankit 1 riser default.fcc Unit Set: Field-Density Date/Time: Mon May 07 00:51:11 2012

FCC Reactor: Reactor Section

DESIGN

Configuration

Number of Risers	Midpoint Injection	Regenerator Type
1 *	No	One-stage

Geometry

Riser

Length	(ft)	105.0 *	
Top Diameter	(ft)	2.461 *	
Bottom Diameter	(ft)	2.461 *	
Injection Point	(ft)	105.0 *	

Regenerator

		Regenerator	
Dense Bed Height	(ft)	14.76 *	
Dense Bed Diameter	(ft)	24.93 *	
Dilute Phase Diameter	(ft)	24.93 *	
Interface Diameter	(ft)	24.93 *	
Cyclone Inlet Height	(ft)	49.21 *	
Cyclone Inlet Diameter	(ft)	7.546 *	
Cyclone Outlet Diameter	(ft)	4.265 *	

Stripper

Height	(ft)	26.25 *
Diameter	(ft)	9.843 *
Annulus Diameter	(ft)	4.265 *

Riser Termination Zone

Length	(ft)	3.281 *
Outer Diameter	(ft)	13.12 *

Heat Loss by Zone

	Heat Loss (Btu/hr)
Riser Heat Loss	0.0000 *
Regenerator Dense Bed Heat Loss	0.0000 *
Regenerator Dilute Phase Heat Loss	0.0000 *
Regenerator Flue Heat Loss	0.0000 *
Reactor Heat Loss	0.0000 *
Reactor Stripper Heat Loss	0.0000 *

FEED DATA

Library

Available Feed Types
Vacuum Gas Oil
Generic Feed
Hydrotreated Atmospheric Resid
Heavy Coker Gas Oil

Hyprotech Ltd. Aspen HYSYS Version 2006.5 (21.0.0.6924) Page 1 of 4

Licensed to: LEGENDS * Specified by user.

aspentech	LEGENDS Calgary, Alberta CANADA	Case Name:	ankit 1 riser default.fcc
		Unit Set:	Field-Density
		Date/Time:	Mon May 07 00:51:11 2012

FCC Reactor: Reactor Section (continued)

CATALYST

Library

Available Catalysts
Conquest 95

Blend

Base Catalyst Blend and Composition

	Weight Fraction	Zeolite	Alumina	Rare Earth
Conquest 95	1.0000 *	24.39	39.69	12.01
Total	1.0000	24.39	39.69	12.01

ZSM 5 Additive

Selectivity	Standard
ZSM-5 per unit mass of base blend	0.0000 *

Heat Capacities

Catalyst Heat Capacity	(Btu/lb-F)	0.2627 *
Coke Heat Capacity	(Btu/lb-F)	0.3989 *

Activity

Feed Metals

	Vanadium (ppmwt)	Nickel (ppmwt)	Sodium (ppmwt)	Iron (ppmwt)	Copper (ppmwt)
Feed-1	0.0000	0.2603	0.0000	0.0000	0.0000
Total	0.0000	0.2603	0.0000	0.0000	0.0000
Bias	-5.608e-002	0.2230	-5.234e-002	-0.1122	-1.495e-003
Equilibrium Catalyst	750.0 *	500.0 *	2800 *	4000 *	20.00 *

Other

Fresh Make Up Rate	(lb/hr)	50.44
Equilibrium MAT	(%)	68.00 *

OPERATION

Feeds

Feed Conditions

Feed	Volume Flow (barrel/day)	Mass Flow (lb/hr)	Temperature (F)	Pressure (psia)	Location
Feed-1	5.010e+004 *	6.746e+005	710.3 *	43.51 *	Riser

Total Feed

	Riser		
Fresh Feed Volume (barrel/day)	5.010e+004		
Fresh Feed Mass (lb/hr)	6.746e+005		
Total Feed Volume (barrel/day)	5.010e+004		
Total Feed Mass (lb/hr)	6.746e+005		
Total Feed Preheat Duty (Btu/hr)	0.0000 *		
Total Feed Temperature (F)	710.3		

Total Feed Summary

		Total
Fresh Feed Volume	(barrel/day)	5.010e+004
Fresh Feed Mass	(lb/hr)	6.746e+005
Total Feed Volume	(barrel/day)	5.010e+004
Total Feed Mass	(lb/hr)	6.746e+005
Total Feed Preheat Duty	(Btu/hr)	0.0000

Dispersion Steam

	Riser	

Hyprotech Ltd. Aspen HYSYS Version 2006.5 (21.0.0.6924) Page 2 of 4

Licensed to: LEGENDS * Specified by user.

LEGENDS
Calgary, Alberta
CANADA

Case Name:	ankit 1 riser default.fcc
Unit Set:	Field-Density
Date/Time:	Mon May 07 00:51:11 2012

FCC Reactor: Reactor Section (continued)

Feeds (continued)

Dispersion Steam (continued)			
Steam Mass (lb/hr)	1.332e+004		
Steam to Total Feed Ratio	2.000e-002 *		
Steam Temperature (F)	392.0 *		
Steam Pressure (psia)	145.0 *		

Riser/Reactor

Riser Temperature Control		
Riser Outlet Temperature	(F)	942.6
Reactor Plenum Temperature	(F)	929.4 *
Catalyst Circulation Rate	(lb/hr)	3.846e+006
Cat/Oil Ratio		5.770
Reactor Stripping Zone		
Stripping Steam Mass Rate	(lb/hr)	1.160e+004
Stripping Steam Temperature	(F)	392.0 *
Stripping Steam Pressure	(psia)	145.0 *
Ratio to Catalyst Circulation Rate		3.000 *

Regenerator

	Regenerator		
Dense Bed Temperature (F)	1249		
Cyclone Temperature (F)	1259		
Flue Gas Temperature (F)	1259		
Flue Gas-Dense Bed Delta-T	5.380		
Flue Gas O2 (%)	1.00 *		
Flue Gas CO (%)	0.24		
Flue Gas CO2 (%)	16.49		
Flue Gas CO/CO2 Ratio (%)	0.01		
Carbon on Regen Cat (CRC) (%)	0.09		
Catalyst Cooler Duty (Btu/hr)	0.0000 *		
Dense Bed Bulk Density	540.0 *		
Catalyst Inventory (lb)	2.430e+005		
Air Volume Flow (barrel/day)	1.881e+007		
Air Mass Flow (lb/hr)	3.319e+005		
Enriched O2 Volume (barrel/day)	0.0000 *		
Enriched O2 Mass Flow (lb/hr)	0.0000		
Enriched O2 Pressure (psia)	14.65 *		
Enriched O2 Temperature (F)	212.0 *		
Air Blower Discharge Temp (F)	392.0 *		

Ambient Air Box		
Temperature (F)	Pressure (psia)	Relative Humidity (%)
392.0 *	14.65 *	70.00 *

Stage 1 Conditions

Dense Bed Temperature	
Apparent	
Bias	

CRC	
Apparent	
Bias	

Hyprotech Ltd. Aspen HYSYS Version 2006.5 (21.0.0.6924) Page 3 of 4

LEGENDS Calgary, Alberta CANADA		Case Name:	ankit 1 riser default.fcc
		Unit Set:	Field-Density
		Date/Time:	Mon May 07 00:51:11 2012

FCC Reactor: Reactor Section (continued)

Pressure Control

Reactor Pressure	(psia)	49.31 *
Regenerator Stage 1 Pressure	(psia)	53.66
Regenerator Stage 2 Pressure		
Regenerator Stage 2 - Reactor Pressure Difference	(psi)	4.351 *
Regenerator Stage 2 - Riser Pressure Difference	(psi)	-4.875

Solver Options

Iteration Limits	
Maximum Iterations	Minimum Iterations
20.00 *	0.0000 *
Convergence Tolerance	
Residual	1.000e-006 *
Variable Scaling Parameter	
On/Off Switch	On
Failure Recovery Action	
Action	Revert to the previous results
Creep Step Parameters	
On/Off Switch	Off
Iterations	10.00 *
Step Size	0.1000 *
SQP Hessian Parameters	
Initialization	Normal
Scaling Factor	1.000 *
Updates stored	10.00 *
Linear Search Parameters	
Algorithm	Normal
Step Control	Normal
Step Control Iterations	0.0000 *

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b) Dual Riser

LEGENDS Calgary, Alberta CANADA	Case Name: ankit 2 riser default.fcc Unit Set: Field-Density Date/Time: Mon May 07 00:57:51 2012

FCC Reactor: Reactor Section

DESIGN

Configuration

Number of Risers	Midpoint Injection	Regenerator Type
2 *	No	One-stage

Geometry

Riser			
Length	(ft)	105.0 *	105.0 *
Top Diameter	(ft)	2.461 *	2.461 *
Bottom Diameter	(ft)	2.461 *	2.461 *
Injection Point	(ft)	105.0 *	105.0 *

Regenerator		
		Regenerator
Dense Bed Height	(ft)	14.76 *
Dense Bed Diameter	(ft)	24.93 *
Dilute Phase Diameter	(ft)	24.93 *
Interface Diameter	(ft)	24.93 *
Cyclone Inlet Height	(ft)	49.21 *
Cyclone Inlet Diameter	(ft)	7.546 *
Cyclone Outlet Diameter	(ft)	4.265 *

Stripper		
Height	(ft)	26.25 *
Diameter	(ft)	9.843 *
Annulus Diameter	(ft)	4.265 *

Riser Termination Zone		
Length	(ft)	3.281 *
Outer Diameter	(ft)	13.12 *

Heat Loss by Zone

	Heat Loss (Btu/hr)
Riser 1 Heat Loss	0.0000 *
Riser 2 Heat Loss	0.0000 *
Regenerator Dense Bed Heat Loss	0.0000 *
Regenerator Dilute Phase Heat Loss	0.0000 *
Regenerator Flue Heat Loss	0.0000 *
Reactor Heat Loss	0.0000 *
Reactor Stripper Heat Loss	0.0000 *

FEED DATA

Library

Available Feed Types
Vacuum Gas Oil
Generic Feed
Heavy Coker Gas Oil
Hydrotreated Atmospheric Resid

Hyprotech Ltd. | Aspen HYSYS Version 2006.5 (21.0.0.6924) | Page 1 of 4

LEGENDS Calgary, Alberta CANADA	Case Name:	ankit 2 riser default.fcc
	Unit Set:	Field-Density
	Date/Time:	Mon May 07 00:57:51 2012

FCC Reactor: Reactor Section (continued)

CATALYST

Library

Available Catalysts
Conquest 95

Blend

Base Catalyst Blend and Composition

	Weight Fraction	Zeolite	Alumina	Rare Earth
Conquest 95	1.0000 *	24.39	39.69	12.01
Total	1.0000	24.39	39.69	12.01

ZSM 5 Additive

Selectivity	Standard
ZSM-5 per unit mass of base blend	0.0000 *

Heat Capacities

Catalyst Heat Capacity	(Btu/lb-F)	0.2627 *
Coke Heat Capacity	(Btu/lb-F)	0.3989 *

Activity

Feed Metals

	Vanadium (ppmwt)	Nickel (ppmwt)	Sodium (ppmwt)	Iron (ppmwt)	Copper (ppmwt)
Feed-1	0.0000	0.2603	0.0000	0.0000	0.0000
Total	0.0000	0.2603	0.0000	0.0000	0.0000
Bias	-3.966	-2.384	-3.702	-7.933	-0.1058
Equilibrium Catalyst	750.0 *	500.0 *	2800 *	4000 *	20.00 *

Other

Fresh Make Up Rate	(lb/hr)	3568
Equilibrium MAT	(%)	68.00 *

OPERATION

Feeds

Feed Conditions

Feed	Volume Flow (barrel/day)	Mass Flow (lb/hr)	Temperature (F)	Pressure (psia)	Location
Feed-1	5.010e+004	6.746e+005 *	710.3 *	43.51 *	Split

Total Feed

	Riser 1	Riser 2	
Fresh Feed Volume (barrel/day)	2.645e+004	2.365e+004	
Fresh Feed Mass (lb/hr)	3.562e+005	3.184e+005	
Total Feed Volume (barrel/day)	2.645e+004	2.365e+004	
Total Feed Mass (lb/hr)	3.562e+005	3.184e+005	
Total Feed Preheat Duty (Btu/hr)	0.0000 *	0.0000 *	
Total Feed Temperature (F)	710.3	710.3	

Total Feed Summary

		Total
Fresh Feed Volume	(barrel/day)	5.010e+004
Fresh Feed Mass	(lb/hr)	6.746e+005
Total Feed Volume	(barrel/day)	5.010e+004
Total Feed Mass	(lb/hr)	6.746e+005
Total Feed Preheat Duty	(Btu/hr)	0.0000

Dispersion Steam

	Riser 1	Riser 2	

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LEGENDS Calgary, Alberta CANADA	Case Name: ankit 2 riser default.fcc
	Unit Set: Field-Density
	Date/Time: Mon May 07 00:57:51 2012

FCC Reactor: Reactor Section (continued)

Feeds (continued)

Dispersion Steam (continued)			
Steam Mass (lb/hr)	1.065e+004	9519	
Steam to Total Feed Ratio	3.000e-002 *	3.000e-002 *	
Steam Temperature (F)	392.0 *	392.0 *	
Steam Pressure (psia)	145.0 *	145.0 *	

Riser/Reactor

Riser Temperature Control		
Riser Outlet Temperature	(F)	935.0 *
Reactor Plenum Temperature	(F)	929.4
Catalyst Circulation Rate	(lb/hr)	4.276e+006
Cat/Oil Ratio		6.278
Reactor Stripping Zone		
Stripping Steam Mass Rate	(lb/hr)	1.282e+004
Stripping Steam Temperature	(F)	392.0 *
Stripping Steam Pressure	(psia)	145.0 *
Ratio to Catalyst Circulation Rate		3.000 *

Regenerator

	Regenerator		
Dense Bed Temperature (F)	1243		
Cyclone Temperature (F)	1253		
Flue Gas Temperature (F)	1253		
Flue Gas-Dense Bed Delta-T	5.751		
Flue Gas O2 (%)	1.00 *		
Flue Gas CO (%)	0.14		
Flue Gas CO2 (%)	16.51		
Flue Gas CO/CO2 Ratio (%)	0.01		
Carbon on Regen Cat (CRC) (%)	0.15		
Catalyst Cooler Duty (Btu/hr)	0.0000 *		
Dense Bed Bulk Density	400.0 *		
Catalyst Inventory (lb)	1.799e+005		
Air Volume Flow (barrel/day)	2.013e+007		
Air Mass Flow (lb/hr)	3.566e+005		
Enriched O2 Volume (barrel/day)	0.0000 *		
Enriched O2 Mass Flow (lb/hr)	0.0000		
Enriched O2 Pressure (psia)	14.65 *		
Enriched O2 Temperature (F)	212.0 *		
Air Blower Discharge Temp (F)	284.0 *		

Ambient Air Box		
Temperature (F)	Pressure (psia)	Relative Humidity (%)
284.0 *	14.65 *	90.00 *

Stage 1 Conditions	
Dense Bed Temperature	
Apparent	
Bias	
CRC	
Apparent	
Bias	

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LEGENDS
Calgary, Alberta
CANADA

Case Name:	ankit 2 riser default.fcc
Unit Set:	Field-Density
Date/Time:	Mon May 07 00:57:51 2012

FCC Reactor: Reactor Section (continued)

Pressure Control

Reactor Pressure	(psia)	23.93 *
Regenerator Stage 1 Pressure	(psia)	34.81 *
Regenerator Stage 2 Pressure		
Regenerator Stage 2 - Reactor Pressure Difference	(psi)	10.88
Regenerator Stage 2 - Riser Pressure Difference	(psi)	7.989

Solver Options

Iteration Limits	
Maximum Iterations	Minimum Iterations
20.00 *	0.0000 *
Convergence Tolerance	
Residual	1.000e-006 *
Variable Scaling Parameter	
On/Off Switch	On
Failure Recovery Action	
Action	Revert to the previous results
Creep Step Parameters	
On/Off Switch	Off
Iterations	10.00 *
Step Size	0.1000 *
SQP Hessian Parameters	
Initialization	Normal
Scaling Factor	1.000 *
Updates stored	10.00 *
Linear Search Parameters	
Algorithm	Normal
Step Control	Normal
Step Control Iterations	0.0000 *

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c) Riser with two stage regenerator

aspentech	LEGENDS Calgary, Alberta CANADA	
Case Name:	C:\Program Files\AspenTech\Aspen HYSYS 2006\Template\one_riser.fc	
Unit Set:	SI	
Date/Time:	Sun May 13 15:04:55 2012	

FCC Reactor: Reactor Section

DESIGN

Configuration

Number of Risers	Midpoint Injection	Regenerator Type
1 *	No	Two-stage, Flue Gas in Series

Geometry

Riser			
Length	(m)	32.00 *	
Top Diameter	(m)	1.000 *	
Bottom Diameter	(m)	1.000 *	
Injection Point	(m)	32.00 *	

Regenerator			
		Stage 1	Stage 2
Dense Bed Height	(m)	4.500 *	4.500 *
Dense Bed Diameter	(m)	7.600 *	7.600 *
Dilute Phase Diameter	(m)	7.600 *	7.600 *
Interface Diameter	(m)	7.600 *	7.600 *
Cyclone Inlet Height	(m)	15.00 *	15.00 *
Cyclone Inlet Diameter	(m)	2.300 *	2.300 *
Cyclone Outlet Diameter	(m)	1.300 *	1.300 *

Stripper		
Height	(m)	8.000 *
Diameter	(m)	3.000 *
Annulus Diameter	(m)	1.300 *

Riser Termination Zone		
Length	(m)	1.000 *
Outer Diameter	(m)	4.000 *

Heat Loss by Zone

	Heat Loss (kJ/h)
Riser Heat Loss	0.0000 *
Stage 1 Dense Bed Heat Loss	0.0000 *
Stage 1 Dilute Phase Heat Loss	0.0000 *
Stage 1 Flue Heat Loss	0.0000 *
Stage 2 Dense Bed Heat Loss	0.0000 *
Stage 2 Dilute Phase Heat Loss	0.0000 *
Stage 2 Flue Heat Loss	0.0000 *
Reactor Heat Loss	0.0000 *
Reactor Stripper Heat Loss	0.0000 *

FEED DATA

Library

Available Feed Types
Vacuum Gas Oil
Feed Type - 1

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aspentech	LEGENDS Calgary, Alberta CANADA	Case Name: C:\Program Files\AspenTech\Aspen HYSYS 2006\Template\one_riser.fc Unit Set: SI Date/Time: Sun May 13 15:04:55 2012

FCC Reactor: Reactor Section (continued)

CATALYST

Library

Available Catalysts
A/F-3

Blend

Base Catalyst Blend and Composition

	Weight Fraction	Zeolite	Alumina	Rare Earth
A/F-3	1.0000 *	26.69	37.20	3.746e-002
Total	1.0000	26.69	37.20	3.746e-002

ZSM 5 Additive

Selectivity		Standard
ZSM-5 per unit mass of base blend		0.0000 *

Heat Capacities

Catalyst Heat Capacity	(kJ/kg-C)	1.100 *
Coke Heat Capacity	(kJ/kg-C)	1.670 *

Activity

Feed Metals

	Vanadium (ppmwt)	Nickel (ppmwt)	Sodium (ppmwt)	Iron (ppmwt)	Copper (ppmwt)
Feed-1	0.0000	0.2606	0.0000	0.0000	0.0000
Total	0.0000	0.2606	0.0000	0.0000	0.0000
Bias	-0.2470	0.1616	-0.4152	-0.5247	-6.587e-003
Equilibrium Catalyst	750.0 *	500.0 *	2800 *	4000 *	20.00 *

Other

Fresh Make Up Rate	(kg/h)	85.83 *
Equilibrium MAT	(%)	68.04

OPERATION

Feeds

Feed Conditions

Feed	Volume Flow (m3/h)	Mass Flow (kg/h)	Temperature (C)	Pressure (kPa)	Location
Feed-1	331.9	3.060e+005 *	376.9 *	300.0 *	Riser

Total Feed

		Riser		
Fresh Feed Volume	(m3/h)	331.9		
Fresh Feed Mass	(kg/h)	3.060e+005		
Total Feed Volume	(m3/h)	331.9		
Total Feed Mass	(kg/h)	3.060e+005		
Total Feed Preheat Duty	(kJ/h)	0.0000 *		
Total Feed Temperature	(C)	376.9		

Total Feed Summary

		Total
Fresh Feed Volume	(m3/h)	331.9
Fresh Feed Mass	(kg/h)	3.060e+005
Total Feed Volume	(m3/h)	331.9
Total Feed Mass	(kg/h)	3.060e+005
Total Feed Preheat Duty	(kJ/h)	0.0000

Dispersion Steam

	Riser	

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aspentech — LEGENDS, Calgary, Alberta, CANADA	Case Name:	C:\Program Files\AspenTech\Aspen HYSYS 2006\Templates\one_riser.fcc
	Unit Set:	SI
	Date/Time:	Sun May 13 15:04:55 2012

FCC Reactor: Reactor Section (continued)

Feeds (continued)

Dispersion Steam (continued)

Steam Mass (kg/h)	6039 *		
Steam to Total Feed Ratio	1.592e-002		
Steam Temperature (C)	200.0 *		
Steam Pressure (kPa)	1000 *		

Riser/Reactor

Riser Temperature Control

Riser Outlet Temperature	(C)	511.1 *
Reactor Plenum Temperature	(C)	505.0
Catalyst Circulation Rate	(kg/h)	1.713e+006
Cat/Oil Ratio		5.341

Reactor Stripping Zone

Stripping Steam Mass Rate	(kg/h)	5034 *
Stripping Steam Temperature	(C)	200.0 *
Stripping Steam Pressure	(kPa)	1000 *
Ratio to Catalyst Circulation Rate		2.838

Regenerator

	Stage 1	Stage 2	
Dense Bed Temperature (C)	753.6	754.3	
Cyclone Temperature (C)	773.4	754.3	
Flue Gas Temperature (C)	773.4	754.3	
Flue Gas-Dense Bed Delta-T	19.79	-6.496e-002	
Flue Gas O2 (%)	9.69	29.94	
Flue Gas CO (%)	0.09	-1.58	
Flue Gas CO2 (%)	10.72	-9.34	
Flue Gas CO/CO2 Ratio (%)	0.01 *	0.19	
Carbon on Regen Cat (CRC) (%)	-0.08	-0.09	
Catalyst Cooler Duty (kJ/h)	0.0000 *	0.0000 *	
Dense Bed Bulk Density	540.0 *	540.0 *	
Catalyst Inventory (kg)	1.088e+005	1.054e+005	
Air Volume Flow (m3/h)	1.325e+005 *	1.325e+005	
Air Mass Flow (kg/h)	1.601e+005	1.601e+005 *	
Enriched O2 Volume Flow (m3/h)	9605	0.0000	
Enriched O2 Mass Flow (kg/h)	1.298e+004	0.0000 *	
Enriched O2 Pressure (kPa)	101.0 *	101.0 *	
Enriched O2 Temperature (C)	100.0 *	100.0 *	
Air Blower Discharge Temp (C)	200.0 *	200.0 *	

Ambient Air Box

Temperature (C)	Pressure (kPa)	Relative Humidity (%)
200.0 *	101.0 *	70.00 *

Stage 1 Conditions

Dense Bed Temperature

Apparent	(C)	1455
Bias	(C)	701.6 *

CRC

Apparent	(%)	-0.01
Bias	(%)	0.07 *

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LEGENDS
Calgary, Alberta
CANADA

Case Name:	C:\Program Files\AspenTech\Aspen HYSYS 2006\Template\one_riser.fcc
Unit Set:	SI
Date/Time:	Sun May 13 15:04:55 2012

FCC Reactor: Reactor Section (continued)

Pressure Control

Reactor Pressure	(kPa)	340.0 *
Regenerator Stage 1 Pressure	(kPa)	370.0 *
Regenerator Stage 2 Pressure	(kPa)	30.00 *
Regenerator Stage 2 - Reactor Pressure Difference	(kPa)	-310.0
Regenerator Stage 2 - Riser Pressure Difference	(kPa)	-319.3

Solver Options

Iteration Limits	
Maximum Iterations	Minimum Iterations
40.00 *	0.0000 *
Convergence Tolerance	
Residual	1.000e-006 *
Variable Scaling Parameter	
On/Off Switch	On
Failure Recovery Action	
Action	Revert to the previous results
Creep Step Parameters	
On/Off Switch	Off
Iterations	10.00 *
Step Size	0.1000 *
SQP Hessian Parameters	
Initialization	Normal
Scaling Factor	1.000 *
Updates stored	10.00 *
Linear Search Parameters	
Algorithm	Normal
Step Control	Normal
Step Control Iterations	0.0000 *

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d) A/F 3 catalyst

FCC Catalyst Name	A/F-3				2M1Butene	1.058146
Description	Akzo A/F-3				C2Pentene	0.938267
Created	Oct-20	2003	17:24 17:24:55		T2Pentene	0.957186
Modified	Oct-20	2003	17:24 17:24:55		Cyclopentene	1.046789
Manufacturer	Akzo				Isoprene	0.958755
Kinetic Coke	1.045989				Benzene	1.5625
Feed Coke	1.166873				Metals H2	1.563636
Stripping Eff.	0.999811				Heat Of Rxn.	0
Metals Coke	1.057143				Bot. Cracking	-0.03785
Methane	1.307692				Fresh MAT	76.05
Ethylene	1.489796				HT Deact.	1.006145
Ethane	1.121951				Met. Deact.	0.611945
Propylene	1.351955				LN RON	2.412
Propane	1.517483				LN MON	1.194
IC4	1.27598				LN Nap.	-0.34
Total C4=	1.318519				LN Olefins	7.28
N Butane	1.051095				LN Aromatics	1.155
IC5	1.235693				LCO SPGR	-0.00837
Total C5=	1.38799				CSO SPGR	-0.0091
NC5	1.017909				SOx	1.037847
IC4=	1.189059				HN RON	2.377714
1Butene	0.943844				HN MON	1.211143
C2Butene	0.947135				HN Nap.	-0.895
Butadiene	1.398742				HN Olefins	1.337143
Cyclopentane	0.793549				HN Aromatics	7.283571
3M1Butene	1.052484				LN SPGR	0.005483
1Pentene	0.92546				HN SPGR	0.007414

Spare 50	0
ZSA M2/GM	166.8
MSA M2/GM	174.8
Zeolite(Wt%)	26.694407
Alumina(Wt%)	37.2
ZRE(Wt%)	0.037461
Sodium(ppm)	1600
Nickel(ppm)	0
Vanadium(ppm)	0
Copper(ppm)	0
Iron(ppm)	2400
ZSM5 LN RON	0
ZSM5 LN MON	0
ZSM5 HN RON	0
ZSM5 HN MON	0
Price	0
Spare 66	0
Spare 67	0
Spare 68	0
Spare 69	0
Spare 70	0

e) Conquest 95 catalyst used in FCC

FCC Catalyst Name	Conquest 95					
Description	Akzo Conquest 95					
Created	Oct-20	2003	17:40 17:40:42		2M1Butene	1
Modified	Oct-20	2003	17:40 17:40:42		C2Pentene	1
Manufacturer	Akzo				T2Pentene	1
Kinetic Coke	1				Cyclopentene	1
Feed Coke	1				Isoprene	1
Stripping Eff.	1				Benzene	1
Metals Coke	1				Metals H2	1
Methane	1				Heat Of Rxn.	0
Ethylene	1				Bot. Cracking	0
Ethane	1				Fresh MAT	80.8
Propylene	1				HT Deact.	0.5
Propane	1				Met. Deact.	0.5
IC4	1				LN RON	0
Total C4=	1				LN MON	0
N Butane	1				LN Nap.	0
IC5	1				LN Olefins	0
Total C5=	1				LN Aromatics	0
NC5	1				LCO SPGR	0
IC4=	1				CSO SPGR	0
1Butene	1				SOx	1
C2Butene	1				HN RON	0
Butadiene	1				HN MON	0
Cyclopentane	1				HN Nap.	0
3M1Butene	1				HN Olefins	0
1Pentene	1				HN Aromatics	0

LN SPGR	0
HN SPGR	0
Spare 50	0
ZSA M2/GM	141.7
MSA M2/GM	183.3
Zeolite(Wt%)	24.38689
Alumina(Wt%)	39.69
ZRE(Wt%)	12.01465
Sodium(ppm)	2100
Nickel(ppm)	0
Vanadium(ppm)	0
Copper(ppm)	0
Iron(ppm)	2500
ZSM5 LN RON	0
ZSM5 LN MON	0
ZSM5 HN RON	0
ZSM5 HN MON	0
Price	0
Spare 66	0
Spare 67	0
Spare 68	0
Spare 69	0
Spare 70	0